



**ATTACHMENTS TO
AMENDMENT AND RESPONSE TO OFFICE ACTION OF JANUARY 14, 2004**

1. U.S. Patent No. 6,103,206
2. U.S. Patent No. 4,844,837
3. U.S. Patent No. 6,682,711
4. "The Lost Cabin Gas Plant Expansion - a Unique Challenge" presented at the Laurance Reid Gas Conditioning Conference, Feb. 27 - Mar. 1, 2000 in Norman, OK (publisher: University of Oklahoma, Oklahoma City, OK).
5. "1993 NPRA Question and Answer Session on Refining and Petrochemical Technology," proceedings of the National Petroleum Refiners Association (NPRA) held in October, 1993, Dallas, Texas (publisher: Gerald L. Farrar & Assocs., Tulsa OK).

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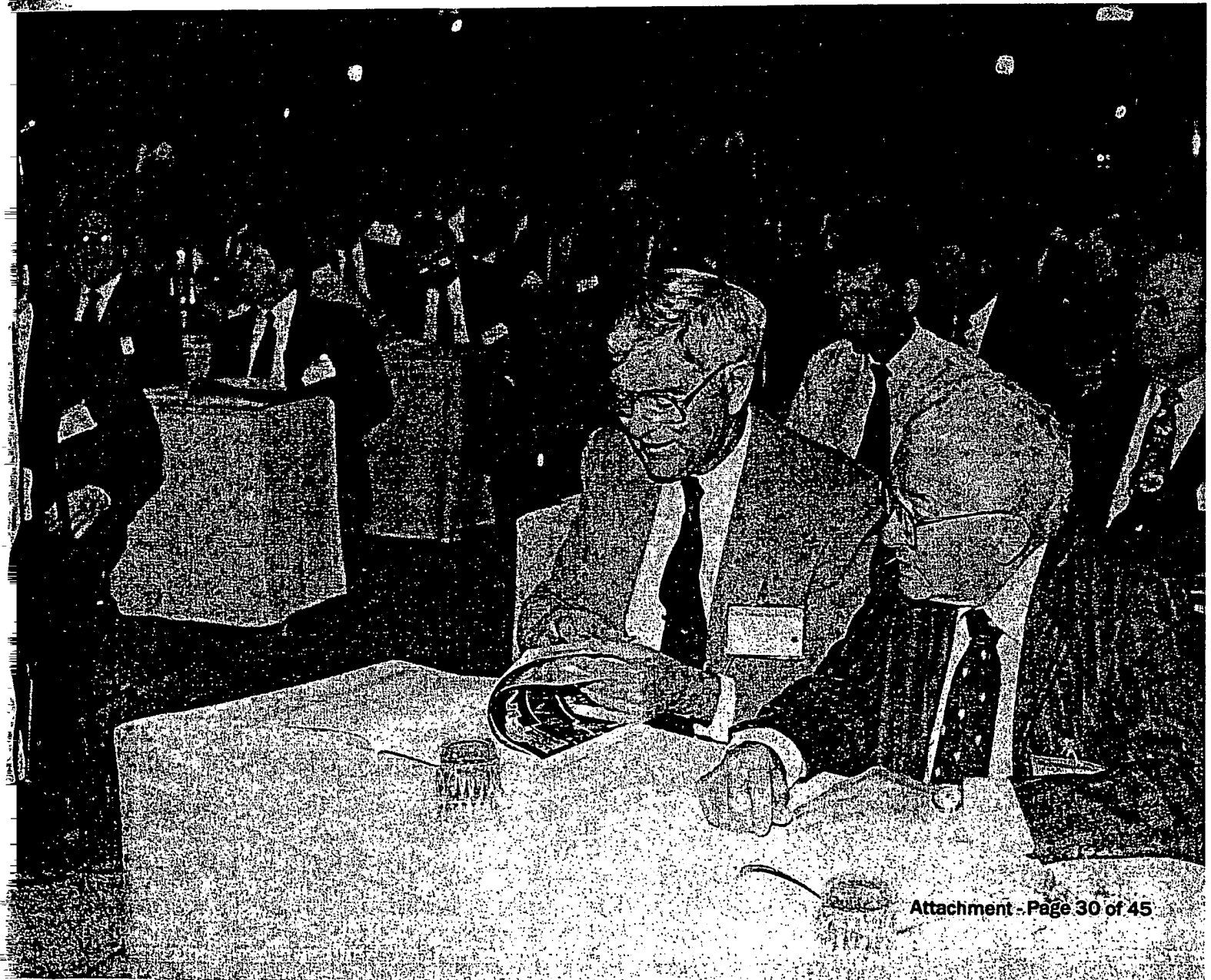
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1993

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vessel and reduce the loading of the hydrogen purification portion of the plant. ICI currently operates two isothermal medium temperature shifts in two of its ammonia plants in Severenside, England since May 1988. Methanol formation is not a concern with the isothermal arrangement because the condensate from the exit gas of the medium temperature shift is converted back to process steam. The higher outlet temperatures that are achieved in the adiabatic situation result in equilibrium methanol levels of only a few parts per million.

HENRY JONES (Howe-Baker Engineers, Inc.):

We have not designed any plants with medium temperature shifts, but would like to ask anybody who is operating one: what type of results do you get with methanol formation in the medium temperature shift, compared, say, to a low or high temperature shift?

HIGGINS:

No answer.

QUESTION 9:

What is the maximum sulfur content for H₂ plant feeds that can be desulfurized by hydrogenation followed by ZnO beds?

SLOAN:

There is no clear limit on sulfur removal by zinc oxide beds. Generally the design is set, based upon changing the zinc oxide on a one year cycle or greater. If the flows are low, the sulfur content can be high and not use up the zinc oxide within the one year cycle. On the other hand, if the flows are high, the sulfur content must be lower.

As a rule of thumb, if the natural gas is of pipeline quality, or the sulfur content does not exceed 40 to 50 ppmv, then zinc oxide can be used. If the sulfur content gets much higher, generally some form of feedstock pretreatment is economically justified.

PAULES:

There is no maximum feed gas sulfur content for a zinc oxide bed, as long as minimum space velocity criteria are followed. However, the more sulfur in the feed, the faster you will spend the zinc oxide, and typically the larger the bed size will need to be.

JUNO:

The limiting factor for sulfur content of H₂ plant feeds would be the amount of sulfur that is unconverted and remains to be reacted in the zinc oxide beds. The sizing of these beds reflects the amount of spent zinc oxide disposal that one wants to contend with and the change out time that would be required. Generally, feed levels above 200 ppm sulfur to the zinc oxide beds tend to be uneconomic. With this sulfur limit set for the desulfurization section outlet, the desulfurizer feed sulfur is set by reactor volume and H₂S cleanup section that is economic for the feed gas that is desired to be processed. One caution is that if the feed contains a large volume of COS, it will be very difficult to remove this sulfur compound with the conventional desulfurization and zinc oxide beds.

Zinc oxide systems are typically designed for 6 months to 2 years operating life depending on the size of the plant and the sulfur level. This type of system by itself is economically used up to 100 to 200 ppm sulfur before the beds become too costly. At higher concentrations of sulfur in the feed gas, a lean amine system is used to absorb the H₂S from the hydrogenated feed down to a level of about 100 ppm and the residual H₂S is then removed by the zinc oxide beds.

HENRY JONES (Howe-Baker Engineers, Inc.):

I tend to agree with Ed Juno that once you get above 200 ppm, you are getting into a situation that is not very economical for zinc oxide only. However, Howe-Baker is currently designing a plant with zinc oxide that has a feedstock sulfur content between 175 and 200 ppm. Obviously, the bed life in this type of plant is not going to be 1 year. You have big beds, and you are going to have to change out the zinc oxide in a shorter time period.

QUESTION 10:

Has anyone successfully used in situ regeneration to regenerate steam methane reformer catalyst that has been contaminated with sulfur?

LAABS:

We have successfully performed this type of regeneration. Sulfur is stripped from the catalyst by running a high steam/carbon ratio. It is very important to bypass the low temperature shift reactor during this regeneration. The sulfur that is stripped will permanently poison the low temperature catalyst. Because of this bypass, we have found it necessary to run the reformed gas to the flare.

RAJGURU:

Our investigation indicates that in situ regeneration has been tried on occasion, with little or no success. We believe that a reliable and adequate sulfur removal system upstream of the reformer pays for itself by extending the catalyst and tube life. In situ regeneration also requires isolation of the reformer from the shift reactor to protect the shift catalyst.

SLOAN:

Sulfur contaminated steam methane reformer catalyst is not necessarily a serious problem. If the source of sulfur contamination is discontinued and the contamination of the reformer catalyst is not severe, the sulfur will be removed through normal operation. It would be advisable to operate with a higher steam to carbon ratio than normal. In some extreme cases, the sulfur has been burned off using air. Occasional sulfur contamination of the reforming catalyst is easily reversed through normal operation combined with higher steam to carbon ratios to accelerate the sulfur removal rate. Many operators have used these techniques to remove sulfur in reformers. It is important to note, however, the sulfur will be carried downstream as it is removed from the reforming catalyst.

T. WILLIAMS:

We did have a minor sulfur contamination problem, and we did a high temperature steam strip. We were losing about 1 to 2% hydrogen purity, which we did recover.

KEN CHLAPIK (ICI Katalco):

Many of our customers have successfully steamed both alkalized and nonalkalized reforming catalysts and have recovered activity by doing so. Much of the activity recovery will depend on the age of the catalyst and the amount of sulfur that has been incorporated on the catalyst. The stability of the catalyst support will also affect the success of the steaming. The stable support systems used in ICI Katalco reforming catalysts allow effective in situ regenerations to occur. I reference the ICI Katalco paper "Safe Start-up and Operation of Steam Reformers" by Dr. Ben J. Cromarty presented at the 1992 AIChE Safety in Ammonia Plants and Related Facilities Symposium in San Antonio TX. It is important to note that steaming will cause some sintering of the nickel sites on the catalyst also affecting the activity of the catalyst. We normally recommend a steaming period of 6 to 12 hours at normal reformer operating conditions to achieve this kind of recovery.

ANDERS NIELSEN (Haldor Topsoe A/S):

In principle it should be possible to regenerate the catalyst by decreasing the sulfur in the feed back to normal. However, while you can put sulfur on very fast, it only comes off at equilibrium level, and diffusion in the pore system further limits the rate.

When the catalyst is regenerated by steaming, S is converted to H_2S and SO_2 . The success of the regeneration in steam depends on the catalyst. If the catalyst is promoted with sodium or potassium, it is nearly unaffected since alkali sulfates are formed.

Reference is made to an article by J. Rostrup-Nielsen describing the chemistry and to a successful industrial regeneration in an article by Andriano and Sinaga. Reference: J. Rostrup Nielsen, *Catalysis* 5, 1984, 101-103. In Andriano and O.G. Sinaga *Ammonia Plant Saf.* 32, 1992, 232-235.

QUESTION 11:

What techniques are currently utilized to reliably predict remaining catalyst life during operation of hydrogen production units?

BRIERLEY:

Temperature profiles in our shift reactors and methanators are trended regularly, along with biweekly checks of CO or CO_2 slip. At least twice a year, operating data and laboratory data are compiled and forwarded to the catalyst vendors for comparative analysis. Their results are compared with our in-house approach to equilibrium calculations.

JUNO:

This answer was obtained from KTI. They basically design for a maximum slip of 0.1 ppm sulfur from the zinc oxide drums. They also design the methane slip from the reformer to be between 3 and 6%, and fired duty is increased to meet design methane slip. End of run for the reformer catalyst is based on the maximum tubeskin temperature.

They normally design the high temperature shift reactor for a maximum inlet temperature of 660 °F, and then design

the low temperature shift for a maximum CO slip of 0.5 volume percent.

LAABS:

We agree with the previous remarks.

CHARLES S. MCCOY (Consultant):

The very concept of "approach to equilibrium" as a measure of steam reformer catalyst deactivation is suspect; in fact, in steam naphtha reforming it simply does not work. Furthermore, analytical imprecision introduces errors in the calculation which net out larger than the alleged catalyst life span. Thus in the particular case of steam-reforming catalyst, asking the supplier whether the catalyst is spent seems a bit like putting the fox in charge of the hen house.

KEN CHLAPIK (ICI Katalco):

ICI Katalco offers simulation capabilities to our hydrogen plant customers in order to do this type of analysis. We strongly encourage them to provide their data to us for these services, but also encourage them to do their own trending of operating parameters around the units. Trending is very important to see subtle changes that are occurring in the performance of the unit before something major happens to require a premature shutdown or a loss of hydrogen purity. Trending allows the removal of systematic errors that occur with analysis and data gathering. Ultimately, a working relationship between the hydrogen plant operator and the catalyst supplier can achieve reliable catalyst life predictions as well as optimize the performance of the catalysts.

BRIERLEY:

I initially suspected that the vendor who supplied the catalyst would suggest that it would last forever, and the other vendors would suggest immediate replacement. But the vendors have been coming back with very close, nearly identical results, and have always been fairly conservative from our point of view, in terms of running the plant. We treasure their input.

QUESTION 12:

How do you handle hydrogen plant process condensate? Can it be degassed and reused as boiler feedwater? What tests are needed to determine water quality? Is degassing done in the hydrogen plant or the refinery utilities area? Are there other uses for this condensate?

BRIERLEY:

Two of our hydrogen plants operate at a steam/carbon ratio of about 6 to 1. Process condensate from these units is steam stripped in a condensate stripper and then treated with a liquid-neutralizing amine-hydroquinone blend to neutralize CO_2 and scavenge oxygen. The stripped condensate is mixed with makeup condensate and used as boiler feedwater.

A single liquid phosphate-dispersant blend is added to the steam drum as part of our congruent phosphate pH treatment program. Alkalinity, pH, phosphate, and conductivity tests are conducted each shift.

THE LOST CABIN GAS PLANT EXPANSION – A UNIQUE CHALLENGE

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ABSTRACT

The Lost Cabin Gas Plant was placed into operation by Louisiana Land and Exploration, a wholly owned subsidiary of Burlington Resources, in the spring of 1995 at a feed rate of 50 MMSCFD. The plant underwent a major two-phase expansion in 1999. The addition of a third well into the Madden Deep Reservoir required a plant capacity increase to 133 MMSCFD. The first expansion phase, completed in the fall of 1998, was to de-bottleneck the plant increasing its capacity to 66.5 MMSCFD. The second phase, completed in the fall of 1999, installed a duplicate of the de-bottlenecked facility.

Due to the addition of the third well, the feed gas that contains 20% CO₂ and 12% H₂S and colloidal sulfur enters the plant at 250°F, 50°F warmer than the original design. The higher temperature taxes the inlet cooling capacity and allows more elemental sulfur to be contained in the gas phase. This sulfur progressively drops out of the gas phase upon pressure reduction and/or cooling, providing an increased challenge in the design of the modifications to the inlet gas cooling equipment. These design challenges will be discussed. Other modifications, which were determined following a high rate performance test of the facility, include changing packing type in the Selexol CO₂ Absorber and using supplemental oxygen in the SRU Reaction Furnace. The primary goal was to develop modifications that would maximize the capacity of the facility while minimizing the requirement for the installation of new equipment. In order to eliminate the need for an additional facility shutdown, several major modifications required for the second phase of the expansion were installed during the first phase.

Operating experience with the de-bottlenecked facility will also be presented along with discussion of the equipment sharing plan between the de-bottlenecked plant and the new expansion equipment.

THE LOST CABIN GAS PLANT EXPANSION – A UNIQUE CHALLENGE

1. PROJECT LOCATION, GAS RESERVES, DRILLING

Louisiana Land and Exploration Company (“LL&E” when mentioned later), a wholly-owned subsidiary of Burlington Resources Inc (“Burlington”), is progressively developing gas reserves from the Madden Deep Reservoir. The processing plants are at Lost Cabin, about 90 miles west of Casper, Wyoming and 3 to 5 miles from the wellheads in the Madden field (see Site photo, Fig 1). Site elevation is 5600 ft above sea level.

Wells are drilled to about 25,000 ft and flow some 45 MM SCFD per well. Well hole diameters range from 26” (to 1600 ft) to 6.5” at full depth. These sizes require larger drill pipe and bottom hole assemblies than normal. The temperature at total depth is 435°F. This high wellbore temperature, and water-sensitive shale formations, required drilling with oil-based mud. State and Federal inspectors, including the Bureau of Land Management (“BLM”) applied a high level of scrutiny. Burlington’s plan for handling oil-based mud and cuttings was approved, as well as the plan for ultimate reclamation of well sites. Oil-based mud is drained from cuttings, reclaimed, and reused in the drilling process. BLM inspections have been positive.

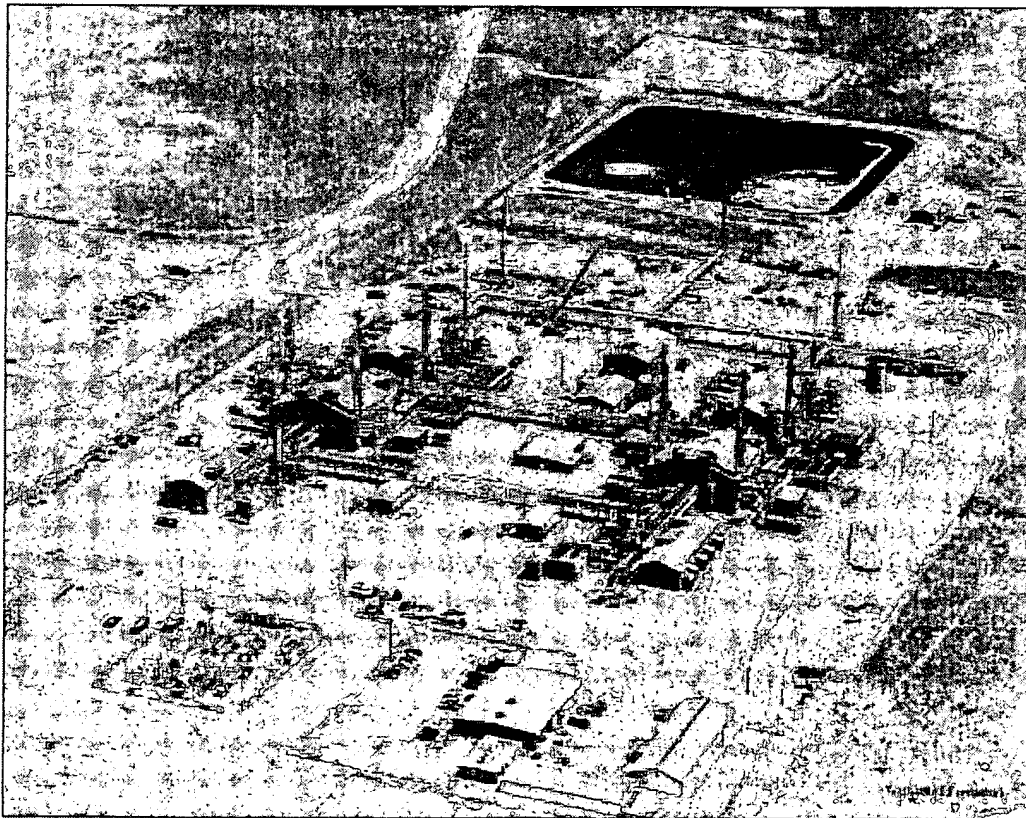


Fig 1. Aerial Photograph, Lost Cabin Site, July 1999

2. PROJECT TIMELINE and PARTICIPANTS

The project was a long time in fruition but, since authorization of Train 1, development activities have proceeded rapidly. LL&E described the original Train 1 as "a 50 MMSCFD pilot plant". Jack Brewster's "Gas Processors Report" applauded the first plant startup in 1995, likening the development as comparable to Exxon's at Mobile Bay. The overall project timeline, depicting completion of Train 1, de-bottlenecking of that train, completion of Train 2, and current work on Train 3, is shown in Table 1.

Table 1 – Project Timeline

- August 1985 Bighorn 1-5 well completed
- Spring 1988 Bighorn 2-3 well completed
- August 1991 DEQ Permit to construct
- August 1993 Start construction of Train 1, 50 MMSCFD initial capacity
- March 1995 Train 1 Startup
- Spring 1997 Bighorn 4-36 well completed
- April 1998 Begin construction of Train 1 (expansion to 66.5 MMSCFD) and of duplicate Train 2
- September 1998 Initiate Turnaround on Train 1
- November 1998 Restart de-bottlenecked Train 1
- July 1999 Train 2 Startup
- Fall 1999 FEED package for larger Train 3

The original processing studies were made in 1986/7 by Mr. Glenn Handwerk, consulting engineer, who continues to contribute technical support to operations and to the process design for Trains 1 and 3. The Arrington Corporation further developed the preliminary designs prior to the award of an engineering contract, and subsequent turnkey contract, for Train 1 to Raytheon Engineers and Constructors ("Raytheon"). Raytheon was later awarded a turnkey contract for Train 2, and is currently developing the Front End Engineering Design (FEED) package for Train 3. The Arrington Corporation has also supported Burlington in permit-related activities, operations training, and Process Safety Management. Dr. Bruce Craig of Metcorr has consulted on materials selection throughout.

3. RAW PRODUCED GAS and WATER

Table 2 summarizes the principal characteristics of the raw fluids. Gas temperatures, at the facilities inlet, have risen as production rates have increased. The carbonyl sulfide concentration, in plant inlet gas, has also risen somewhat in the same period.

Table 2 – Raw Gas and Water

- Wellhead temperatures measured close to 300°F
- Designing for 300°F inlet Train 3 as flows increase
- 12+% H₂S, 20-% CO₂, up to 400 ppm COS
- No heavy hydrocarbons
- Gas carries up to 15 bbl water per MMSCF, with variable chloride content
- Elemental sulfur in gas phase, progressively condenses

4. GAS PROCESSING OVERVIEW

The processing block flow diagram is shown in Fig 2.

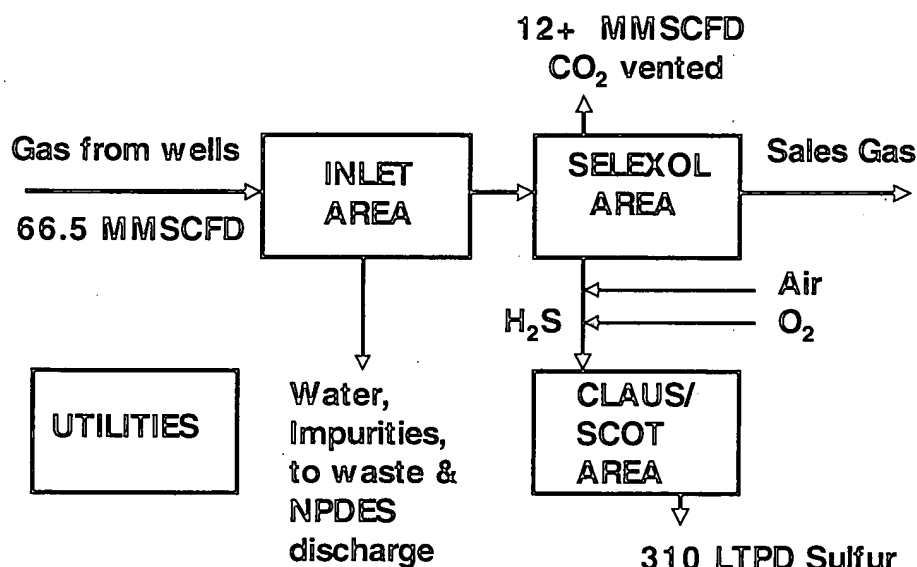


Figure 2 - Lost Cabin Train 2 Process Block Flow Diagram

The high content of acid gases favors using a physical solvent. Selexol is used for selective removal of first H₂S, then CO₂, from the gas. The Selexol process design is a simpler version of the selective process designed by Raytheon, for Exxon's LaBarge plants, in the 1980s. The larger train sizes at LaBarge, and the much higher CO₂/H₂S ratio in raw gas, justified the more complex process used there. The LaBarge design was described in detail in Ref 1, to which the reader is referred for discussion of the design principles of selective Selexol processing. A brief description of the Lost Cabin Selexol process is given later in this paper, coupled with discussion of the unusual corrosion experienced at Lost Cabin. LaBarge and

Lost Cabin are the only reported locations utilizing Selexol for selective removal of H_2S and CO_2 from natural gas.

The use of refrigerated Selexol solvent mandates substantial pre-cooling of the feed gas, and this presents a very difficult processing problem at Lost Cabin. A significant amount of excess elemental sulfur (in addition to the sulfur content of the H_2S) is in gas phase solution in the inlet gas, either as sulfur or as hydrogen polysulfides. The solubility of this excess sulfur decreases with reduction of either gas pressure or gas temperature, resulting in hundreds of pounds per day of sulfur crystals trying to settle somewhere. That amount would make a slurry of <0.1 wt % sulfur, if dispersed in the excess produced water. However, as discussed later, it proved a formidable task to persuade most of the sulfur to deposit in places and in ways that minimize operating and maintenance problems.

The selective Selexol process generates a rich ($\sim 85\%$ H_2S) sulfur plant feed. The original Train 1 sulfur plant was an air-blown Claus unit, de-bottlenecked later with an oxygen-enriched air supplement. Since Burlington decided to build Train 2 as a near-duplicate of Train 1, Train 2 also uses oxygen-assist in the Claus plant. Oxygen-assisted Claus technology has been widely reported elsewhere, but a few details of the Lost Cabin Claus process are provided later in this paper.

The Wyoming Department of Environmental Quality BACT requirements were met with a conventional SCOT unit on the Claus tail gas. SCOT tail gas goes to a Thermal Oxidizer with a waste heat boiler generating 450 PSIG steam.

There is very little good quality natural water available at Lost Cabin. However, because the raw gas is so sour, the SCOT Sour Water Stripper supplies enough good water (polished in an R.O. unit) that a plant water balance is easily maintained. The produced water coming in with the inlet gas is filtered of elemental sulfur, treated in its own Sour Water Stripper, and then discharged under permit. Small amounts of ammonia are currently complicating the sour water stripping process, and that will be discussed later.

5. GAS PROCESSING SPECIFICS

In the following subsections, details are provided of the several process sections in the overall flow scheme described above. Priority is given to those details that are novel, or where operations taught us that design improvements were needed. Processes that are similar to those in many other plants, or which have been well covered already in the literature, are discussed only briefly.

5.1 Inlet Area, Gas Cooling, Water and Sulfur Separation

Gas coming from the pipeline at up to $300^\circ F$ has to be cooled significantly before it can flow to the Selexol unit for three main reasons:

(a) Hot gas to Selexol would greatly increase the cost of solvent refrigeration.

- (b) Hot gas, carrying more elemental sulfur in gas-phase solution, would build up an increasing sulfur load in the Selexol system.
- (c) It is desirable to reduce the quantity of gas-phase water so as to lower the dehydration duty in the Selexol unit.

In 1992, Alberta Sulfur Research Ltd (ASR) provided results of gas analyses they had made of Madden Deep gas samples provided by LL&E. See Fig 3. These data predicted that elemental sulfur in gas-phase solution, in excess of the sulfur in H_2S , would precipitate in the Inlet Area cooling section and pose a formidable equipment fouling problem.

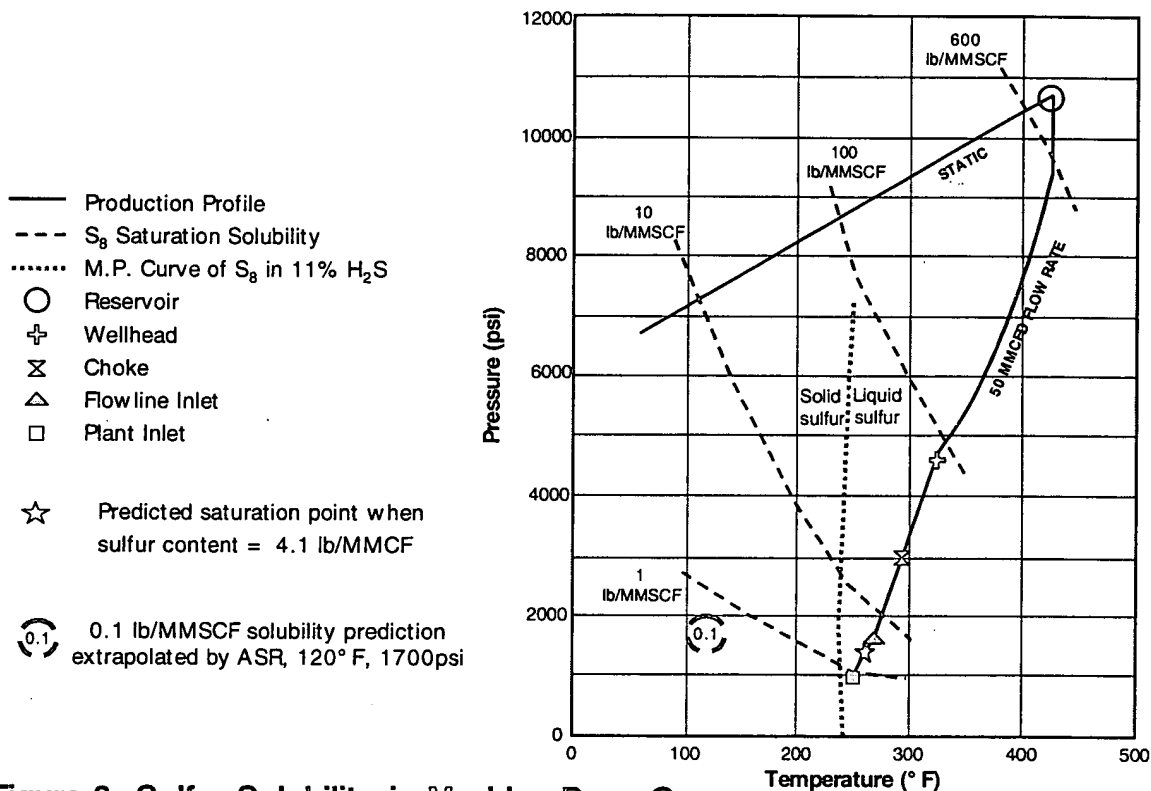


Figure 3: Sulfur Solubility in Madden Deep Gas - simplified from an original diagram by Alberta Sulfur Research Ltd (ASR)

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It was clear that conventional indirect cooling equipment would foul with sulfur very quickly, and a special study for the inlet area was commissioned in 1993. Having successfully used static mixers in co-current gas-liquid flow, for both heat and mass transfer duties in several plants built in the 1980s, Raytheon proposed an inlet cooling system that was (and we think may still be) unique in gas plants. Co-current cooling is, of course, less efficient than countercurrent cooling, but co-current devices can be made highly turbulent and thus less vulnerable to fouling with solids. Fig 4 shows the combination, of turbulent co-current cooling and efficient countercurrent cooling, that has been developed for Lost Cabin gas and that will (with minor modifications) be used again for the new Train 3.

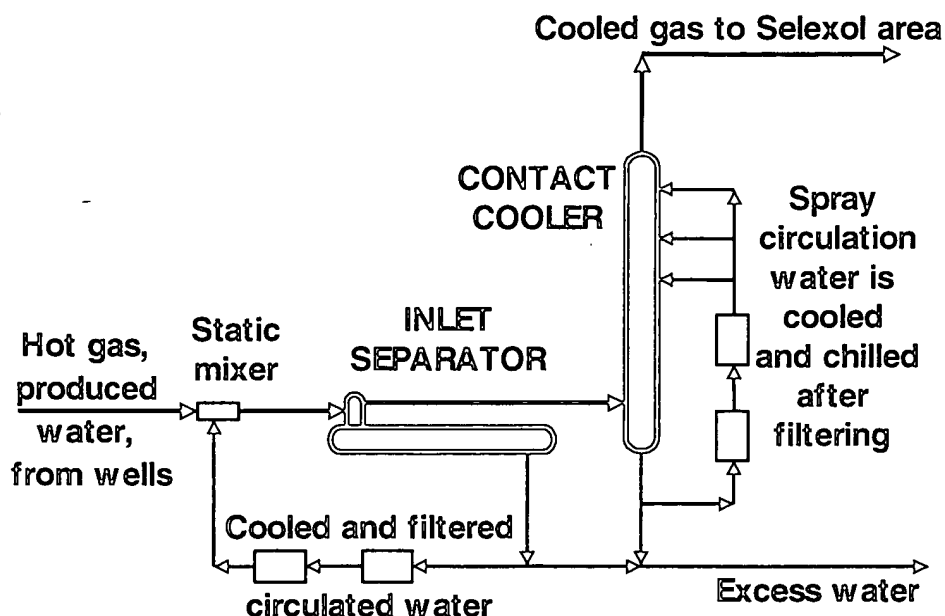


Figure 4 - Lost Cabin Train 2 Inlet Area Block Flow Diagram

The original design for Train 1 was to use water, cooled in air coolers to 110°F, in the static mixer. Water, chilled to 85°F by exchange against gas leaving the Selexol unit, would be used in a trayed Contact Cooler. By these means, it was intended to cool the incoming gas and liquid to 130°F in the Static Mixer, and then to further cool the gas to below 100°F in the Contact Cooler.

The Inlet Separators comprised simple dome separators above small horizontal slugcatchers, all designed as piping rather than pressure vessels. The cooling performance of the Inlet Area equipment was, initially, reasonably satisfactory but problems soon developed with solid sulfur fouling. The chevron vanes at the top of the inlet domes quickly blocked with sulfur and were removed. Produced water carryover then increased the salt content of water in the Contact Cooler loop, in turn threatening salt carryover into Selexol. Most of the elemental sulfur did precipitate from the gas when it cooled in the static mixer but the original plan, to filter the excess produced water only after letdown to a lower pressure, failed. Sulfur rapidly fouled the surfaces of all the cooler tubes, and filters were added to both high-pressure water circulation loops. The carbon steel cooler tubes had corroded under the sulfur deposits. They were replaced with thicker wall carbon steel tubes; filtering the circulating water, and use of corrosion inhibitors and wetting agents, has now brought corrosion rates under control.

More severe sulfur deposition problems were experienced in the Contact Cooler, despite most of the inlet sulfur having been caught in the Inlet Separators. To any sulfur carryover from the Inlet Separators, was added the sulfur precipitating from the gas as it cooled further in the tower. The lower two trays, in particular, fouled rapidly and were replaced with a more turbulent, fouling-resistant design which did not solve the problem. All trays were then removed, and later replaced by water sprays. After trials with different spray types, gas was being cooled to around 100°F upstream of Selexol, with six-month runs between cooler tube clean-outs, but with much time devoted to filter changing.

The nature of the sulfur deposits, found within the plant, varied with location. In the Inlet Separator loop, deposits were yellow-black and crusty. In the Contact Cooler itself, brilliant yellow clusters resembling "hoar frost" several inches thick would be found at shutdowns (see Fig 5). It appears that the gas, cooling in the tower, supersaturates with sulfur with potential to precipitate tiny crystals. Such crystals may need an existing sulfur nucleus to precipitate, or they could form and then dodge the water droplets, as would happen with fume.

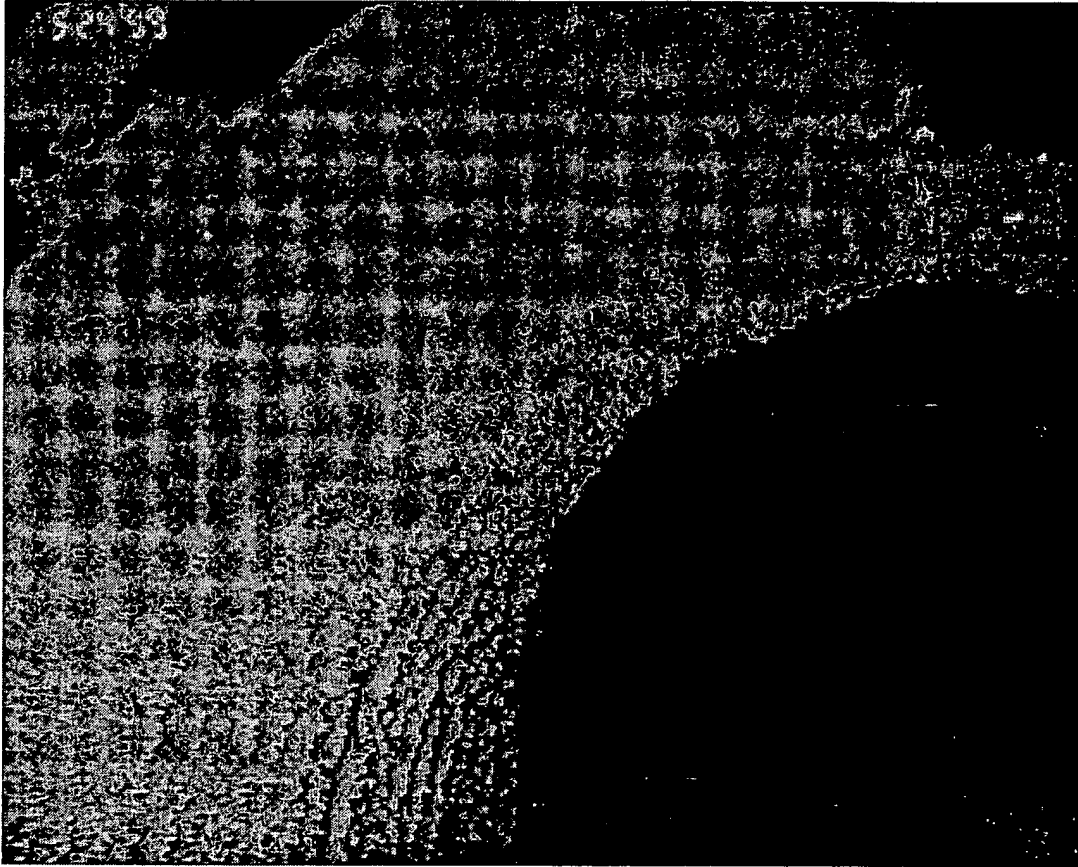


Fig 5. Sulfur Crystals in the top manway of Train 1 Contact Cooler, May 1999

Downstream of the Contact Cooler, just ahead of the Selexol unit, a small filter unit was provided in the original design. Its purpose was to catch small quantities of elemental sulfur carrying over from the top of the Contact Cooler, or precipitating out as a result of line pressure loss. This filter did catch small amounts of sulfur, but was occasionally damaged by slugs of water. A later modification added a centrifugal separator ahead of this filter to prevent water reaching the filter.

In 1998, Train 1 was de-bottlenecked to higher rates, and Train 2 was added in 1999. The water rates to the parallel static mixers were increased, but the original Inlet Separators were not augmented. Once Train 2 started up, the Separators showed two signs of being overloaded:

- (a) More liquid water is carrying over to the Contact Coolers. The centrifugal separator downstream of each Contact Cooler is collecting substantial amounts of water.
 - (b) Only a small fraction of the inlet sulfur is currently being caught in the Inlet Separator loop.
- That recent change requires more discussion, below.

The "warm gas" entering the Contact Cooler is now markedly warmer with increased throughput (currently as warm as 210 °F) but that was expected. Even at this temperature, ASR's 1992 equilibrium data would predict that most of the sulfur in the hot inlet gas should still drop out in the Inlet Separators. But that is not observed. Of three possible causes below, (a) and (b) below seem less likely:

- (a) ASR's predictions of sulfur solubility equilibria, as a function of temperature, were substantially in error, or
- (b) Sulfur crystallizes from the gas, but mostly carries over into the Contact Cooler, or
- (c) The shorter residence time does not now allow sulfur to drop out of gas-phase solution, to the extent that equilibrium data would predict.

A short-term modification is currently being made, to the piping arrangements at one of the existing Inlet Separators, which may soon shed some light on the cause of the sudden change in sulfur separation.

The overloads on the Inlet Areas of Trains 1 and 2 will be relieved, when Train 3 is built, by installing a common new Inlet Area facility that will cool all incoming gas to 200°F. A portion of the gas from that new facility will flow on to Trains 1 and 2, which will then see gas inlet temperatures as low as they were in 1995.

The new Inlet Area will again use the proven combination of turbulent co-current cooling to drop the gas temperature enough to knock out most of the sulfur, followed by countercurrent cooling in a direct contact spray tower to drop the gas temperature further. The larger Train 3 justifies a somewhat more complex Contact Cooler scheme, which is still under development. Data from Trains 1 and 2 allows more precise design of the large new tower for Train 3. The contact cooling process is simple heat transfer without much mass transfer, but is not well modeled by available process simulators that offer only "tray" models. For this purpose, Raytheon developed a new model which builds on work by Fair (Ref 2) and which has been calibrated using data from Trains 1 and 2.

5.2 Selexol Area

Fig 6 shows a block diagram of the gas and solvent flows, and some typical temperatures, in Trains 1 and 2 Selexol units. Readers will find a detailed discussion, of selective sour gas processing with Selexol, in Ref 1. With raw gas containing 12+% H₂S and 20-% CO₂, the selective scheme shown can produce a Claus feed up to 85% H₂S while easily meeting sales gas specs on H₂S, CO₂, and COS. Most of the COS tends to go with the CO₂, which is vented to atmosphere at Lost Cabin. As regulatory pressure to lower COS emissions is increasing, a more sophisticated Selexol system design will be required for Train 3. Raytheon and Burlington have explored several alternates to meet these increased demands, which ranged considerably in their capital and operating costs and their effects on Claus feed concentration. Burlington is currently preparing permit-related proposals to the Wyoming Department of Environmental Quality.



The unusual problem, experienced in the Train 1 Selexol equipment at Lost Cabin, was the onset of serious environmental cracking in stainless steel in the Stripper section in June 1996, after 14 months of operation. The timeline for the cracking problems is in Table 3.

- (a) 304 packing in Stripper beds 1 and 2 (lower) having operated for 14 months prior to the first tower inspection, cracked and crushed one month later.
- (b) The new 316 packing, substituted in beds 1 and 2, failed similarly after only six weeks in service.
- (c) Support grids, "hammered and rung true" during the second inspection, cracked and collapsed six weeks later.

Currently, metallurgical tests are under way, to explore for less costly corrosion resistant alloys for Train 3. The change in surge capacity location, referred to above as an operational improvement, also saves significant capital given these materials choices. The large surge

volume, provided in corrosion-resistant alloy drums at the base of the Strippers on Trains 1 and 2, will be moved for Train 3 to a location where carbon steel is acceptable.

Table 3 – Train 1 Corrosion Timeline

- 3/95 Startup of Train 1
- 5/96 First turnaround. The Selexol unit was opened up for the first time:
 - Stripper top (bed 4), fouled, was replaced with new packing
 - Old 304 SS rings were reported "O.K."
 - No detailed inspection was made, but no corrosion was seen in the Stripper.
- 6/96 Forced Shutdown - Cracking failures in Stripper:
 - Cracked/crushed rings in beds 1 & 2 were replaced in 316.
 - Small cracks found in the solid 316 SS Stripper Surge Drum.
 - Bed Support grids were hammered and "rang true".
- 8/96 Forced Shutdown - major cracking failures
 - 6-week-old 316 rings found cracked, replaced with Inconel.
 - Surge Drum, major cracks, replaced by C.S. drum, lined with corrosion-resistant alloy.
 - Support grids cracked/collapsed, replaced with Hastelloy 625.

In June 1999, clusters of stainless steel packing rings were hung in a Stripper Surge Drum at various levels. They were removed after two months exposure to operating conditions, and have recently been inspected by Dr. Craig, who found no sign of cracking. This renews the debate, begun in 1996, over what combination of causes, whether during operation or during shutdowns, led to the cracking in Train 1. However, until the causes can be more clearly identified, and ways seen to surely prevent their arising again, corrosion resistant alloys will continue to be used in Lost Cabin Selexol strippers.

De-bottlenecking Selexol: Increasing the inlet gas capacity of Train 1 by 33% required much detailed work in the Selexol unit. That work was based on rigorous prior plant performance testing, pushing rates to their limits and measuring flows, compositions and pressure drops. Heat exchange capacity was increased in key areas, solvent distributors were modified, and packing size was increased in the base of the CO₂ Absorber. It was

decided to increase gas recycle rates by operating spare compressors in parallel, rather than investing in more compressors.

The Train 2 Selexol design essentially followed the lines of the de-bottlenecked Train 1, except that the diameter of the CO₂ Absorber was increased, and larger recycle compressor units were purchased.

5.3 Claus and SCOT Areas

The original Claus unit in Train 1 was a conventional three-bed unit with COS hydrolysis catalyst in the first bed, and will not be described here. The conventional SCOT unit design initially included use of caustic for Quench pH control, but Burlington switched to ammonia (a) because they found it easier to control and (b) because it allowed use of the stripped water in their boilers.

Table 4 gives a brief summary of the Claus gas flows, before and after de-bottlenecking, and shows the higher H₂S content in raw gas than was anticipated when Train 1 was first designed. The figures show how an increase of 43% in sulfur tonnage was achieved by oxygen addition, with a relatively small increase in the gas traffic through the Claus plant. Burlington decided to counter the higher pressure drop through the plant by running the two air compressors in parallel during warm weather.

Table 4 - Train 1 Claus Flow Rates, before and after de-bottlenecking

	Original PFD (11.3%)	De-bottlenecked PFD (12.1%)
(raw gas dry % H ₂ S)		
Sulfur product LTPD	215	309
O ₂ , lb/hr, from Air	10,016	10,337
O ₂ , lb/hr, from "oxygen"	nil	4,024
O ₂ , lb/hr, total	10,016	14,362
Claus Tail Gas, lb/hr	54,401	60,985
Claus Tail Gas, mol wt.	25.75	24.8

The SCOT plant is conventional and has generally met the design needs satisfactorily. It was de-bottlenecked by procedures similar to those described earlier for the Selexol plant. Quench Tower trays were changed to two-pass design. The SCOT Sour Water Stripper height was increased. The Train 1 reactor was not modified; Shell advised that the catalyst life would probably drop from ten years to eight. For Train 2, the reactor size was increased.

Lastly, although the problem also appears at the plant inlet, discussion of ammonia in process water streams can be placed here in this "SCOT" section. The only place where ammonia is deliberately added is in the SCOT Quench tower, as mentioned earlier.

However, ammonia concentration in SCOT sour water persists to a surprising degree, even when ammonia additions are small or absent. Further, ammonia is detected at the plant inlet, in produced water from the Inlet Separators. The Inlet Area Produced Water Stripper has the capacity to drive both the ammonia and the H₂S (which are small amounts in terms of mass flow) overhead to the Claus plant. Since the SCOT sour water stripper is operated to prevent ammonia leaving the SCOT/Claus loop in stripped water, the Claus furnace must control the ammonia buildup. The Claus furnace was not designed for ammonia destruction, and the ammonia build-up is making operation of the SCOT sour water stripper difficult.

Currently, an analytical study is under way to determine how much ammonia, if any, is being introduced from the possible sources listed below.

- (a) ammonia in raw inlet gas
- (b) ammonia from degradation of filming amine inhibitor used in the Inlet Area
- (c) ammonia formed, if NOX is formed in the Claus and/or SCOT furnaces and then reduced in the reactors
- (d) ammonia deliberately added to SCOT quench
- (e) ammonia from degradation of minor amine contamination of SCOT quench water, at the high temperatures of the Stripper reboiler

6. CONCLUSIONS

Development of the substantial Madden Deep gas reserves has presented a set of unique challenges to Burlington Resources. These challenges, whether originating from the technical concerns of sulfur deposition in the Inlet Area or from the unknowns of drilling nearly 5 miles below the surface, have been overcome by technical innovation fostered by a team building environment.

The team of owner, contractor, and consultants was able to safely deliver Train 1 on schedule and Train 2 one month ahead of schedule while dealing with the issues presented earlier. The ongoing design efforts on Train 3 were initiated, and have continued to develop, in an environment that benefits from the open discussion of ideas and expression of opinions.

With Trains 1 and 2 having achieved several years of safe operational experience, with economies of scale evident in the larger Train 3 plant, and with savings of over \$10 million in the cost related to the latest well, the economic future of Train 3 looks promising.

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